The velocity is given by equation 4.64:

Thus $$u^2 = 7P_1v_1(1 - w^{0.29})$$
$$= 7 \times 10 \times 10^6 \times 0.0103(1 - w^{0.29})$$

$\therefore$ $$u = 0.849 \times 10^3(1 - w^{0.29})^{0.5} \text{ m/s}$$

Velocity of sound at a pressure of $10 \times 10^6 w$ N/m^2

$$= \sqrt{1.4 \times 10 \times 10^6 w \times 0.0103 w^{-0.71}}$$
$$= 380 w^{0.145} \text{ m/s}$$

$$\text{Mach number} = 0.849 \times 10^3(1 - w^{0.29})^{0.5}/(380 w^{0.145}) = 2.23(w^{-0.29} - 1)^{0.5} \quad (2)$$

(2) *The non-isentropic compression across the shock wave.* The velocity downstream from the shock wave (suffix s) is given by equation 4.54:

$$u_s = u_1[(\gamma - 1)Ma_1^2 + 2]/[Ma_1^2(\gamma + 1)]$$
$$= 0.849 \times 10^3(1 - w^{0.29})^{0.5}(0.4 \times 4.97(w^{-0.29} - 1) + 2)/[4.97(w^{-0.29} - 1) \times 2.4]$$
$$= 141(1 - w^{0.29})^{-0.5} \text{ m/s} \quad (3)$$

The pressure downstream from the shock wave P_s is given by equation 4.53:

$$P_s/10 \times 10^6 w = [2\gamma Ma_1^2 - (\gamma - 1)/(\gamma + 1)]$$

Substituting from equation 2:

$$P_s = 56.3 w(w^{-0.29} - 1) \times 10^6 \text{ N/m}^2 \quad (4)$$

(3) *The isentropic expansion of the gas to atmospheric pressure.* The gas now expands isentropically from P_s to P_a (= 101.3 kN/m^2) and the flow area increases from A to the full bore of 0.07 m^2. Denoting conditions at the outlet by suffix a, from equation 4.64:

$$u_a^2 - u_s^2 = 7P_sv_s[1 - (P_a/P_s)^{0.25}] \quad (5)$$
$$u_a/v_a = 85.7/0.07 = 1224 \text{ kg/m}^2\text{ s} \quad (6)$$
$$u_s/v_s = 85.7/A \text{ kg/m}^2\text{ s} \quad (7)$$
$$v_a/v_s = (P_a/P_s)^{-0.71} \quad (8)$$

Equations 1, 3 to 8, involving seven unknowns, can be solved by trial and error methods to give $w = 0.0057$.

Thus the pressure upstream from the shock wave is:

$$10 \times 10^6 \times 0.0057 = 0.057 \times 10^6 \text{ N/m}^2$$

or $$\underline{\underline{57 \text{ kN/m}^2}}$$

Mach Number, from equation 2: $\underline{\underline{= 4.15}}$

Pressure downstream from shock wave P_s, from equation 4:

$$= \underline{\underline{1165 \text{ kN/m}^2}}$$

4.3 TWO-PHASE FLOW

In this section some features of the flow of a two-phase fluid formed from a gas or vapour and a liquid will be discussed. Steady flow of such fluids occurs in oil and gas lines and in many

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4.3.1. Flow Pattern

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situations in the process industries. The flow of water and steam in boilers and evaporators and the flow of partially condensed vapour–liquid mixtures are further illustrations.

Because of the presence of the two phases there are considerable complications in describing and quantifying the nature of the flow compared with conditions with a single phase. The lack of knowledge of the velocities at a point in the individual phases makes it impossible to give any real picture of the velocity distribution. In most cases the gas phase, which may be flowing with a much greater velocity than the liquid, continuously accelerates the liquid thus involving a transfer of energy. Either phase may be in *streamline* or in *turbulent* flow, but the most important case is that in which both phases are turbulent. The criterion for streamline or turbulent flow of a phase is whether the Reynolds number for its flow at the same rate on its own is less or greater than 1000–2000. This distinction is to some extent arbitrary in that injection of a gas into a liquid initially in streamline flow may result in turbulence developing.

The more important factors in two-phase flow which a designer must seek to understand and be able to evaluate are the stability of the flow, the likelihood of the flow causing erosion problems, the relative volumes of the two phases present in the pipeline (the *holdup*) and the pressure drop caused by the presence of an additional phase. For a detailed treatment of the subject, the reader is referred to works by Govier and Aziz[6], by Chisholm[7] and by Hewitt[8].

4.3.1. Flow Pattern

The principal flow patterns which are observed in gas-liquid flow in both horizontal and vertical pipes are illustrated in Fig. 4.7 and their characteristics are briefly described in Table 4.1. The distinction between any two flow patterns is far from clear-cut, especially in the case of bubble, plug and slug flows, which are sometimes lumped together and called "intermittent" flow. The conditions under which each flow pattern exists can conveniently be represented on a "flow pattern map" and several groups of workers have produced their own maps[9–13]. Most of the data used for compiling such maps have been obtained for the flow of water and air at near atmospheric temperature and pressure, and scaling factors have been introduced to extend their applicability to other systems. However, bearing in mind the diffuse nature of the boundaries between the regimes and the relatively minor effect of changes in

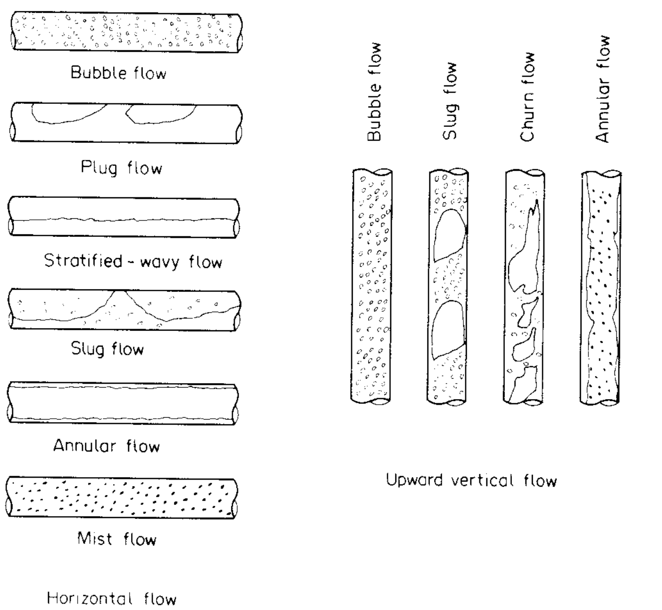

FIG. 4.7. Flow patterns in two-phase flow.

physical properties, such a refinement does not appear to be justified. The flow pattern map for horizontal flow illustrated in Fig. 4.8 which has been prepared by Chhabra and Richardson[14] is based on those previously presented by Mandhane *et al.*[13] and Weisman *et al.*[12] The axes of this diagram are superficial liquid velocity u_L and superficial gas velocity u_G (in each case the volumetric flowrate of the phase divided by the total cross-sectional area of the pipe).

TABLE 4.1. *Flow regimes in horizontal two-phase flow*

Regime	Description	Typical velocities (m/s) Liquid	Typical velocities (m/s) Vapour
Bubble flow[a]	Bubbles of gas dispersed throughout the liquid	1·5–5	0·3–3
Plug flow[a]	Plugs of gas in liquid phase	0·6	<1·0
Stratified flow	Layer of liquid with a layer of gas above	<0·15	0·6–3
Wavy flow	As stratified but with a wavy interface due to higher velocities	<0·3	>5
Slug flow[a]	Slugs of gas in liquid phase	Occurs over a wide range of velocities	
Annular flow[b]	Liquid film on inside walls with gas in centre		>6
Mist flow[b]	Liquid droplets dispersed in gas		>60

([a]) Frequently lumped together as *intermittent* flow
([b]) Sometimes merged as *annular/mist* flow

Slug flow should be avoided when it is necessary to obviate unsteady conditions, and it is desirable to design so that annular flow still persists at loadings down to 50 per cent of the normal flow rates. Even though in many applicants both phases should be turbulent, excessive gas velocity will lead to a high pressure drop, particularly in small pipes. Instability may arise at any point where the phases separate, and in upward vertical flow a high vapour velocity may give rise to instability.

4.3.2. Holdup

Because the gas always flows at a velocity greater than that of the liquid, the *in situ* volumetric fraction of liquid at any point in a pipeline will be greater than the input volume fraction of

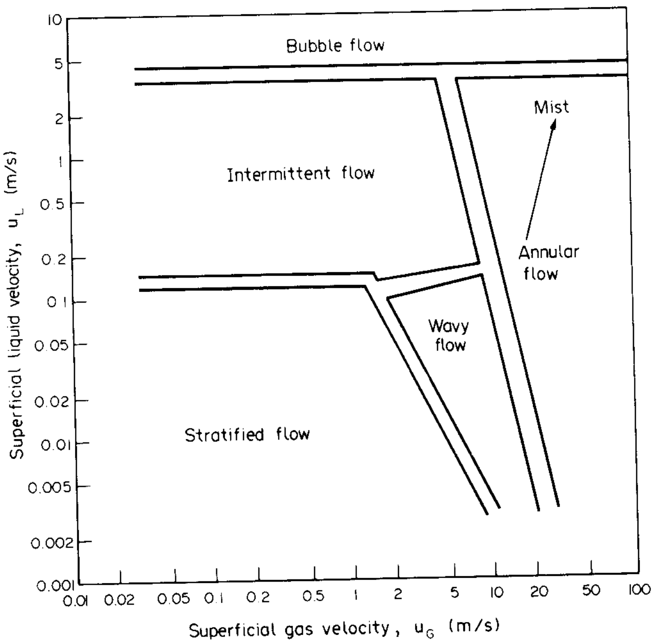

FIG. 4.8. Flow pattern map.

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liquid; furthermore it will progressively change along the length of the pipe as a result of expansion of the gas.

There have been several experimental studies of two-phase flow in which the holdup has been measured, either directly or indirectly. The direct method of measurement involves suddenly isolating a section of the pipe by means of quick-acting valves and then determining the quantity of liquid trapped[15,16]. Such methods are cumbersome and are subject to errors arising from the fact that the valves cannot operate instantaneously. Typical of the indirect methods is that in which the pipe cross-section is scanned by γ-rays and the holdup is determined from the extent of their attenuation[17,18].

Lockhart and Martinelli[19] expressed holdup in terms of a parameter X, characteristic of the relative flowrates of liquid and gas, defined as follows:

$$X = \sqrt{\frac{-\Delta P_L}{-\Delta P_G}} \tag{4.66}$$

where $-\Delta P_L$ and $-\Delta P_G$ are the frictional pressure drops which would arise from the flow of the respective phases on their own at the same rates. Their correlation is reproduced in Fig. 4.9. As a result of more recent work it is now generally accepted that the correlation overpredicts values of liquid holdup. Thus Farooqi and Richardson[20], the results of whose work are also shown in Fig. 4.9, have given the following expression for liquid holdup h_L for co-current flow of air and Newtonian liquids in horizontal pipes:

$$\left.\begin{aligned} h_L &= 0{\cdot}186 + 0{\cdot}0191\,X & 1 < X < 5 \\ h_L &= 0{\cdot}143\,X^{0.42} & 5 < X < 50 \\ h_L &= \frac{1}{0{\cdot}97 + 19/X} & 50 < X < 500 \end{aligned}\right\} \tag{4.67}$$

When the liquid exhibits non-Newtonian behaviour, equation 4.67 is applicable only for turbulent flow; for streamline flow of liquid, modifications are necessary, as given by Farooqi and Richardson[20] and by Chhabra, Farooqi and Richardson[21].

A knowledge of holdup is particularly important for vertical flow since the hydrostatic pressure gradient, which is frequently the major component of the total pressure gradient, is directly proportional to liquid holdup. However, in slug flow, the situation is complicated by the fact that liquid which is in the form of an annular film surrounding the gas slug does not contribute to the hydrostatic pressure[22]:

4.3.3. Pressure, Momentum, and Energy Relations

Methods for determining the drop in pressure start with a physical model of the two-phase system, and the analysis is developed as an extension of that used for single-phase flow. In the *separated flow* model the phases are first considered to flow separately; and their combined effect is then examined.

The total pressure gradient in a horizontal pipe, $(-\mathrm{d}P_{TPF}/\mathrm{d}l)$, consists of two components which represent the frictional and the acceleration pressure gradients respectively,

i.e. $$(-\mathrm{d}P_{TPF}/\mathrm{d}l) = (-\mathrm{d}P_f/\mathrm{d}l) + (-\mathrm{d}P_a/\mathrm{d}l) \tag{4.68}$$

A momentum balance for the flow of a two-phase fluid through a horizontal pipe and an energy balance may be written in an expanded form of that applicable to single-phase fluid flow. These equations for two-phase flow cannot be used in practice since the individual phase velocities and local densities are not known. Some simplification is possible if it is assumed that the two phases flow separately in the channel occupying fixed fractions of the total area, but even with this assumption of separated flow regimes, progress is difficult. It is important to note that, as in the case of single-phase flow of a compressible fluid, it is no longer possible to relate the shear stress to the pressure drop in a simple form since the pressure drop now covers both frictional and acceleration losses. The shear at the wall is proportional to the total rate of momentum transfer (arising from friction, acceleration, and potential effects), so that the total drop in pressure $-\Delta P_{TPF}$ is given by:

$$-\Delta P_{TPF} = (-\Delta P_f) + (-\Delta P_a) \tag{4.69}$$

The pressure drop due to acceleration is important in two-phase flow because the gas is normally flowing much faster than the liquid, and therefore as it expands the liquid phase will accelerate with consequent transfer of energy. For flow in a vertical direction, an additional term $-\Delta P_{gravity}$ must be added to the right hand side of equation 4.69 to account for the hydrostatic pressure attributable to the liquid in the pipe, and this may be calculated approximately provided that the liquid holdup is known.

Analytical solutions for the equations of motion are not possible because of the difficulty of specifying the flow pattern and of defining the precise nature of the interaction between the phases. Rapid fluctuations in flow frequently occur and these cannot readily be taken into account. For these reasons, it is necessary for design purposes to use correlations which have been obtained using experimental data. Great care should be taken, however, if these are used outside the limits used in the experimental work.

Practical Methods for Evaluating Drop in Pressure

Probably the most widely used method for estimating the drop in pressure due to friction is that proposed by Lockhart and Martinelli[19] and later modified by Chisholm.[23] This is based on the physical model of separated flow in which each phase is considered separately and then a combined effect formulated. The two-phase pressure drop due to friction $-\Delta P_{TPF}$ is taken as the pressure drop $-\Delta P_L$ or $-\Delta P_G$ that would arise for either phase flowing alone in the pipe at the stated rate, multiplied by some factor Φ_L^2 or Φ_G^2. This factor is presented as a function of the ratio of the individual single-phase pressure drops.

$$-\Delta P_{TPF}/-\Delta P_G = \Phi_G^2 \tag{4.70}$$

$$-\Delta P_{TPF}/-\Delta P_L = \Phi_L^2 \tag{4.71}$$

The relation between Φ_G and Φ_L and X (defined by equation 4.66) is shown in Fig. 4.10, where it is seen that separate curves are given according to the nature of the flow of the two phases. This relation was developed from studies on the flow in small tubes of up to 25 mm diameter with water, oils, and hydrocarbons using air at a pressure of up to 400 kN/m^2. Although, as indicated earlier,

$$Re_L\left(\frac{L'd}{\mu_L}\right) \quad \text{and} \quad Re_G\left(\frac{G'd}{\mu_G}\right)$$

are used as criteria for defining the flow regime, values less than 1000 to 2000 do not necessarily imply that the fluid is in truly viscous flow. Later experimental work showed that the total pressure has an influence and data presented by Griffith[24] should be consulted where pressures are in excess of 3 MN/m^2. Chisholm[23] has developed a relation between Φ_L and X which he puts in the form:

$$\Phi_L^2 = 1 + (C/X) + (1/X^2) \tag{4.72}$$

where C has a value of 20 for turbulent–turbulent flow,
10 for turbulent liquid, streamline gas,
12 for streamline liquid, turbulent gas, and
5 for streamline/streamline flow.

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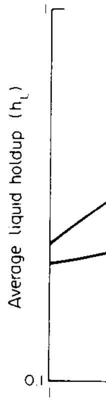

Example 4.5. S
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the pressure drop

Solution. Cross-s

Flow of water

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Density of steam

Flow of steam

equivalent to a

If the densities of the fluids are significantly different from those for water and air at atmospheric temperature and pressure the values of C are somewhat modified.

Chenoweth and Martin[25,26] have presented an alternative method for calculating the drop in pressure. Their method is empirical and based on experiments with pipes of 75 mm and pressures of up to 0·7 MN/m^2. They have plotted the volume fraction of the inlet stream that is liquid as abscissa against the ratio of the two-phase pressure drop to that for liquid flowing at the same volumetric rate as the mixture. An alternative technique has been described by Baroczy.[27] If heat transfer gives rise to evaporation then reference should be made to work by Dukler *et al.*[28]

An illustration of the method of calculation of two-phase pressure drop is included here as Example 4.5.

When the liquid exhibits shear-thinning non-Newtonian characteristics, the above procedures are valid, provided that the liquid flow is turbulent. For streamline flow of liquid, substantial reductions in pressure gradient may result from the injection of gas and this phenomenon is discussed in Volume 2, Chapter 7.

Critical Flow

For the flow of a compressible fluid, conditions of sonic velocity may be reached, and impose limitations on the maximum flowrate for a given upstream pressure. This situation can also occur with two-phase flow, and such critical velocities may sometimes be reached with a drop in pressure not much above 30 per cent of the inlet pressure.

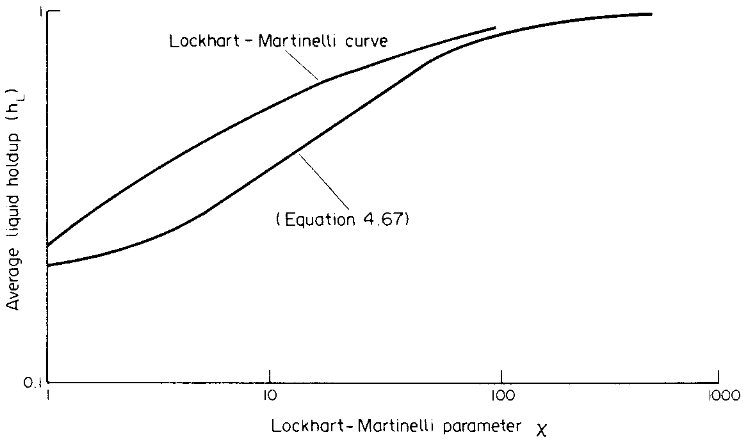

FIG. 4.9. Correlation for average liquid holdup h_L.

Example 4.5. Steam and water flow through a 75 mm i.d. pipe at flowrates of 0·05 and 1·5 kg/s respectively. If the mean temperature and pressure are 330 K and 120 kN/m^2, what is the pressure drop per unit length of pipe assuming adiabatic conditions?

Solution. Cross-sectional area for flow $= \pi(0{\cdot}075)^2/4 = 0{\cdot}00442\ \text{m}^2$

Flow of water $= 1{\cdot}5/1000 = 0{\cdot}0015\ \text{m}^3/\text{s}$

equivalent to a velocity of $0{\cdot}0015/0{\cdot}00442 = 0{\cdot}339$ m/s

Density of steam at 330 K and 120 kN/m^2

$$= (18/22{\cdot}4)(273/330)(120/101{\cdot}3) = 0{\cdot}788\ \text{kg/m}^3$$

Flow of steam $= 0{\cdot}05/0{\cdot}788 = 0{\cdot}0635\ \text{m}^3/\text{s}$

equivalent to a velocity of $0{\cdot}0635/0{\cdot}00442 = 14{\cdot}37$ m/s

Viscosities at 330 K and 120 kN/m^2:

$$\text{steam} = 0{\cdot}0113 \times 10^{-3}\ \text{N s/m}^2; \quad \text{water} = 0{\cdot}52 \times 10^{-3}\ \text{N s/m}^2$$

Therefore $$Re_L = 0{\cdot}075 \times 0{\cdot}339 \times 1000/(0{\cdot}52 \times 10^{-3}) = 4{\cdot}89 \times 10^4$$

$$Re_G = 0{\cdot}075 \times 14{\cdot}37 \times 0{\cdot}788/(0{\cdot}0113 \times 10^{-3}) = 7{\cdot}52 \times 10^4$$

That is, both the gas and liquid are in turbulent flow.

From the friction chart (Fig. 3.7), assuming $e/d = 0{\cdot}00015$:

$$(R/\rho u^2)_L = 0{\cdot}0025 \quad \text{and} \quad (R/\rho u^2)_G = 0{\cdot}0022$$

$\therefore$ From equation 3.15:

$$-\Delta P_L = 4(R/\rho u^2)_L(l/d)(\rho u^2) = 4 \times 0{\cdot}0025(1/0{\cdot}075)(1000 \times 0{\cdot}339^2) = 15{\cdot}32\ (\text{N/m}^2)/\text{m}$$

$$-\Delta P_G = 4 \times 0{\cdot}0022(1/0{\cdot}075)(0{\cdot}778 \times 14{\cdot}37^2) = 18{\cdot}85\ (\text{N/m}^2)/\text{m}$$

$\therefore$ $$-\Delta P_L/-\Delta P_G = 15{\cdot}32/18{\cdot}85 = 0{\cdot}812$$

that is $$X^2 = 0{\cdot}812 \quad \text{and} \quad X = 0{\cdot}901$$

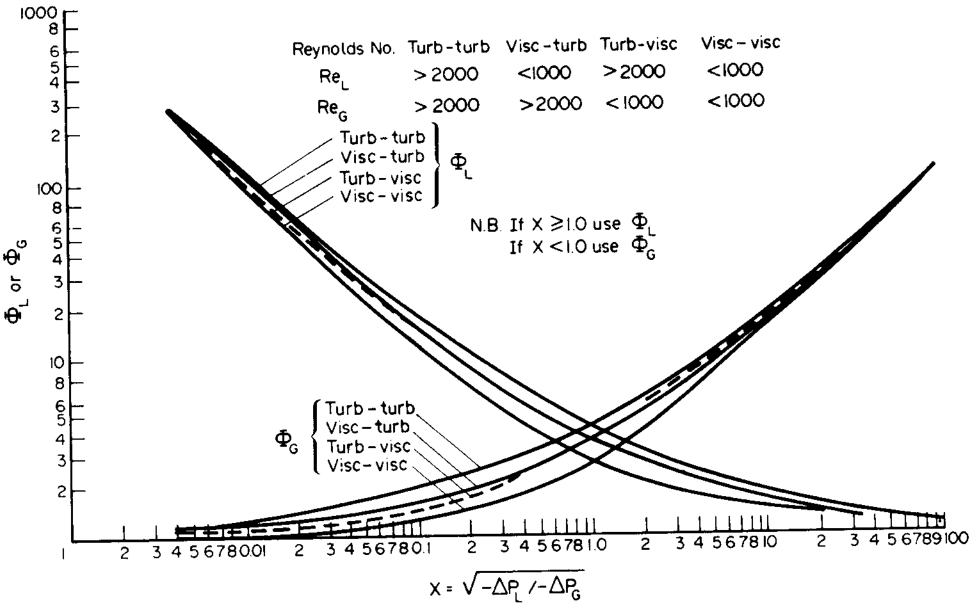

FIG. 4.10. Relationship between Φ and X for two-phase flow.

From Fig. 4.10, for turbulent–turbulent flow,

$$\Phi_L = 4{\cdot}35 \quad \text{and} \quad \Phi_G = 3{\cdot}95$$

Therefore $$-\Delta P_{TPF}/-\Delta P_G = 3{\cdot}95^2 = 15{\cdot}60$$

and $$-\Delta P_{TPF} = 15{\cdot}60 \times 18{\cdot}85 = 294\ (\text{N/m}^2)/\text{m}$$

or $$-\Delta P_{TPF} = 0{\cdot}29\ (\text{kN/m}^2)/\text{m}$$

4.3.4. Erosion

Two-phase systems are often accompanied by erosion, and many empirical relationships have been suggested to avoid the condition. Since high velocities may be desirable to avoid the instability associated with slug flow, there is a danger that any increase in throughput above

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the normal operating condition will lead to a situation where erosion may become a serious possibility.

An indication of the velocity at which erosion becomes significant may be obtained from

$$\rho_m u_m^2 = 15{,}000 \tag{4.73}$$

where ρ_m is the mean density of the two-phase mixture (kg/m^3) and u_m the mean velocity of the two-phase mixture (m/s),

$$\rho_m = (L' + G')/[(L'/\rho_L) + (G'/\rho_G)] \tag{4.67}$$

and

$$u_m = u_L + u_G \tag{4.75}$$

where u_L and u_G are the superficial velocities of the liquid and gas respectively.

It is apparent that some compromise may be essential between the avoidance of both a slug-flow condition and the velocities which are likely to cause erosion conditions.

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